REQ. NO. 8-9114

CONTRACT REPORT NO. OSQ79-00097

FEASIBILITY STUDY
FOR THE PRODUCTION OF
INTERMEDIATE BTU FUEL GAS
FROM COAL AT
EDMONTON, ALBERTA
1977

MICROMEDIA

SNC
TOTTRUP
SERVICES
LTD

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SHELL GASIFICATION STUDY

1.0 INTRODUCTION

This study was commissioned by Shell Canada Resources Limited on behalf of itself and its partners to examine the feasibility of using coal deposits north of Edmonton to provide fuel gas for selected industrial consumers in the Edmonton area, thus freeing natural gas for other users. Partners in this study included the Federal Government of Canada, the Provincial Government of Alberta, Shell Canada Resources Limited, PanCanadian Petroleum Limited, Dow Chemical Canada Limited, Sherritt Gordon Mines Limited, and SNC Tottrup Services Ltd.

It was the belief of the partners that the ultimate feedstock for both the chemical industry and the power industry in Alberta would be coal. It was decided therefore to examine the economic potential of using a particular, well located coal deposit to meet the needs of a selected few industrial customers now totally reliant on natural gas.

Major constraints on the study included:

1. Only commercially proven gasification technology could be used.
2. The lower heating value of the produced gas must be at least 300 BTU's per SCF (11.2 MJ/m^3).
3. The plant size should be as large as the market could reasonably be expected to absorb, but not so large as to make the mine exceed an economic size.
4. The mine and plant should have a 30 year life.

2.0 SUMMARY OF RESULTS

Under certain circumstances, coal can be gasified to produce a fuel gas that will be competitive with natural gas. The most significant requirement, however, is that the price of natural gas must rise substantially.

If we accept the latest projections made by Alberta and Southern Gas Co. Ltd. and Trans Canada Pipelines Ltd. (A.E.R.C.B. Proceeding No. 7700 48), then we can expect that large industrial consumers in Alberta will be paying $2.74/GJ ($2.89/million Btu) delivered in 1985 in 1977 constant dollars. At this price level with a low royalty, the proposed complex can be competitive.

If production from the gasification plant could be used to divert natural gas into the export market, where it could receive a price comparable with frontier gas and LNG, the complex could be an economic venture.

Under the more conservative rules that formed the basis for the study, however, the project would be only marginally economic because the capital cost of current technology is too high. Improvements are necessary if the project is to pass the economic tests of industry.

2.1 Physical Parameters

The plant size was established at 150 billion BTU per day (158.59 billion kJ/d). Two surface mines would have to be developed to supply approximately 18,000 metric tons per day of coal to the plant site. Electric draglines would be used for stripping overburden, with shovels and front end loaders used for mining the coal. The coal would be transported an average distance of five miles in 100 ton trucks. Ash from the process and the boilers would be trucked to the mine site for disposal.

The plant would be sited near the mine, and would use the Lurgi fixed bed gasification system, the Lurgi Rectisol gas purification system, the Stretford sulfur recovery system and the Lurgi Phenolsolvan system for ammonia and phenol recovery.

The gas produced would have a heating value of 424 BTU's per standard cu.ft. (15.8 MJ/m^3). By-products produced include tar, oil, naphtha, crude phenols and ammonium sulfate.

The gas would be delivered by pipeline to the customers' property line at about 22 bar (300 psig). Continuity of supply would be assured by storing gas at 220 bar in salt caverns.

Auxiliary plants required include two 1,300 metric ton per day oxygen plants, two 370,000 kilogram per hour high pressure boilers, a 280 metric ton per day sulphuric acid plant, a 375 metric ton per day ammonium sulfate plant, a large refrigeration plant, as well as water treatment, waste water treatment, cooling water systems, instrument air, etc.

2.0 SUMMARY OF RESULTS - cont'd

2.1 Physical Parameters - cont'd

Approximately 40 megawatts of power would be generated from high pressure steam using back pressure steam turbine generator sets. An additional 15 MW would be produced from 6.9 bar steam using condensing turbines. 24 MW would be purchased for the plant with additional power purchased for mining, gas storage and water pumping.

Approximately 3,000 IGPM of water is required for the plant. This would be removed from the North Saskatchewan River and pumped to a reservoir at the site. All plant and mine runoff would be pumped into this reservoir, as would most waste waters after treatment. Dissolved salt levels in the water would be controlled by demineralizing and injecting the demineralizer effluent and other blowdowns into a disposal well.

Approximately six years would be required from the beginning of development work (additional testing, studies, permit applications, etc.) to the initial production of fuel gas. Delivery of gas to the first customer would occur about 15 weeks after startup, with full capacity delivered approximately 45 weeks later. All gasifiers would be operational in an additional 12 weeks, for a total start-up period of 72 weeks.

2.2 Financial Parameters

The total capital cost for the entire system to mine, convert and deliver energy as fuel gas to the customers' sites is estimated to be approximately $669,389,000 in constant 1977 dollars. This number includes all owners' costs including project management and start-up costs until the end of the first year of operation, but does not include any interest during construction.

Annual operating costs for the plant at full production (with no provisions for the cost of capital) is estimated to be approximately $52,000,000 or about $.88/million BTU of total fuel produced. ($0.83/GJ).

The base case is a joint venture operation with full flow through of tax credits, a 47% tax rate, a 30% capital cost allowance, 100% equity (0.0 Debt/Equity Ratio), no royalty costs, and no value of coal in the ground. The gas price required to produce a D.C.F. rate-of-return of 15% for this case is $2.60/GJ.

The most favourable tax regime is to operate as a joint venture with all tax credits flowing back to the owners, using a 50% capital cost allowance. Even this configuration cannot approach a 15% D.C.F. rate-of-return using 1977 capital cost.

2.0 SUMMARY OF RESULTS - cont'd

2.2 Financial Parameters - cont'd

Sensitivity to capital costs, operating costs, tax rates, capital cost allowance (depreciation), schedule, Debt/Equity Ratios, royalties and operating costs have been tested.

D.C.F. rate-of-return is most sensitive to Debt/Equity Ratios in the joint venture case. This is because the owner's net capital investment becomes so small compared to total accumulated cash flow for the high revenue cases, while interest charges and debt repayment approach gross profits for the low revenue cases.

The 47% tax rate results in a lower gas price than the 41% rate because of the Resource Income Allowance, which more than compensates for the 6% lower tax rate.

To show how royalties might affect the project, they were calculated on gas revenue using the Government of Alberta coal formula. The effect is to increase the gas price by $0.71/GJ at the target 15% D.C.F. rate-of-return.

Each one dollar added to the cost of the coal mined adds about $0.11/GJ to the price of gas.

Each one dollar royalty added to the cost of coal mined adds about $0.15/GJ to the price of gas.

A delay of one year in bringing the project on line adds $0.47/GJ to the Base Case price of gas.

The maximum allowable capital cost allowance (two year write-off) lowers the Base Case gas price by $0.14/GJ.

A drop of 30% in capital cost lowers operating costs by 15% and reduces the gas price to the target level of $1.90/GJ under the most favourable tax regime (50% C.C.A)

TABLE 2.2.1

GAS COSTS TO GIVE 15% D.C.F. ROR $/GJ

TAX RATE			0%	41%	47%		47%	
CAPITAL COST ALLOWANCE				30%	30%	50%	30%	30%
Debt/Equity Ratio	0	S.A.	2.81	3.26	3.20			
		J.V.	-	2.75	2.60	2.46	3.07	3.31
	0.5	S.A.	-	2.83	2.75	-	-	-
		J.V.	-	1.95	1.85	-	-	-
	0.7	S.A.	-	2.71	2.63	-	-	-
		J.V.	-			-	-	-
Debt/Equity Ratio	Venture Type						Year Delay	Royalty

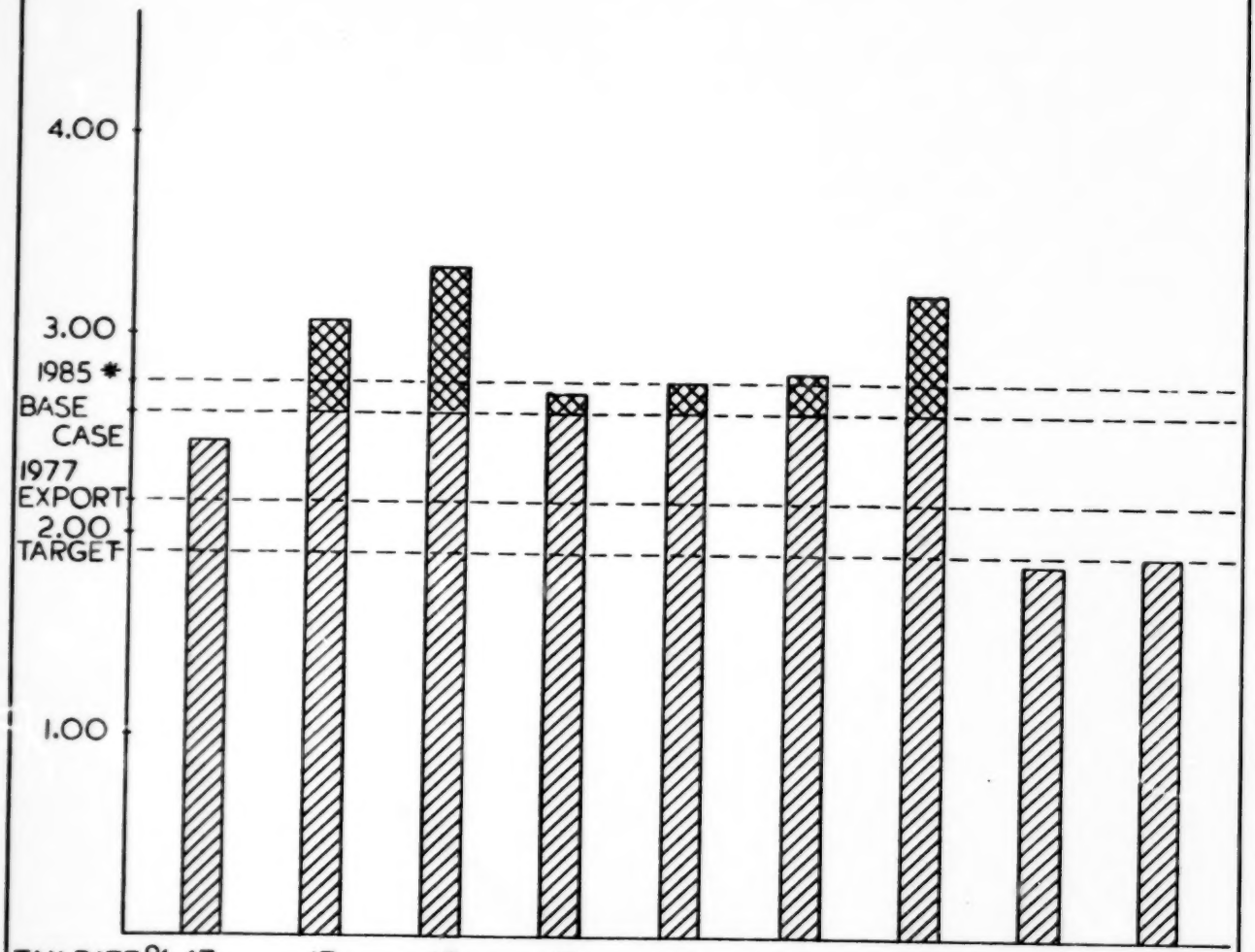
S.A. - Stand Alone

J.V. - Joint Venture

Target Price = $1.90/GJ

snc	FEUILLE DE CALCUL	CLIENT SHELL	CONTR 0249	SUBDIV	ELEMENT	DIV / DEPT	
	DESIGN SHEET	PROJ COAL GASIFICATION	FAIT/MADE WRM	APPR	DATE	PAGE	

SUJET / SUBJECT: GAS PRICE TO RETURN 15% DCF, $/GJ , 100% EQUITY BASIS



TAX RATE, %	47	47	47	47	41	0	47	47	47
C.C.A, %	50	30	30	30	30	0	30	30	50
VENTURE	J.V.	J.V.	J.V.	J.V.	J.V.	S.A.	S.A.	S.A.	J.V.
ROYALTY	0	0	COAL FORMULA	0	0	0	0	0	0
COAL COST	0	0	0	$1.00	0	0	0	0	0
SENSITIVITY	C.C.A.	DELAY	ROYALTY	COAL COST	TAX	TAX	VENTURE TYPE	DEBT (50%)	CAPITAL (-30%)

* 1985 ERCB PROJECTION (IN CONSTANT 1977 DOLLAR) COST TO LARGE INDUSTRIAL CUSTOMERS.

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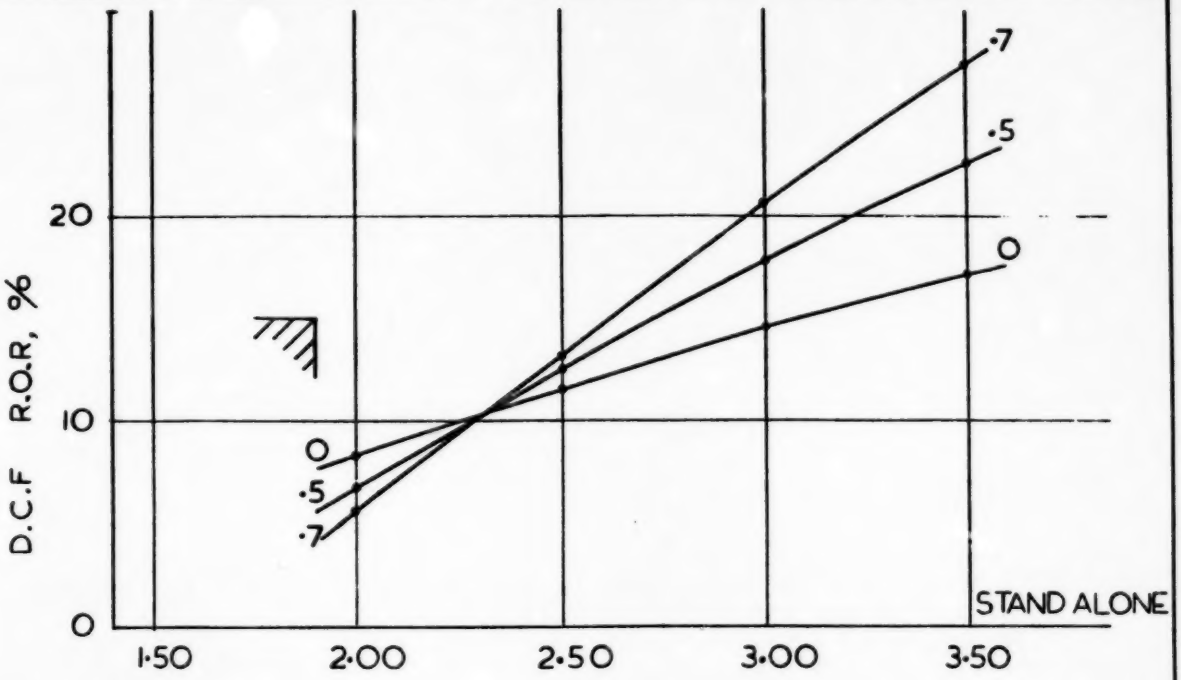
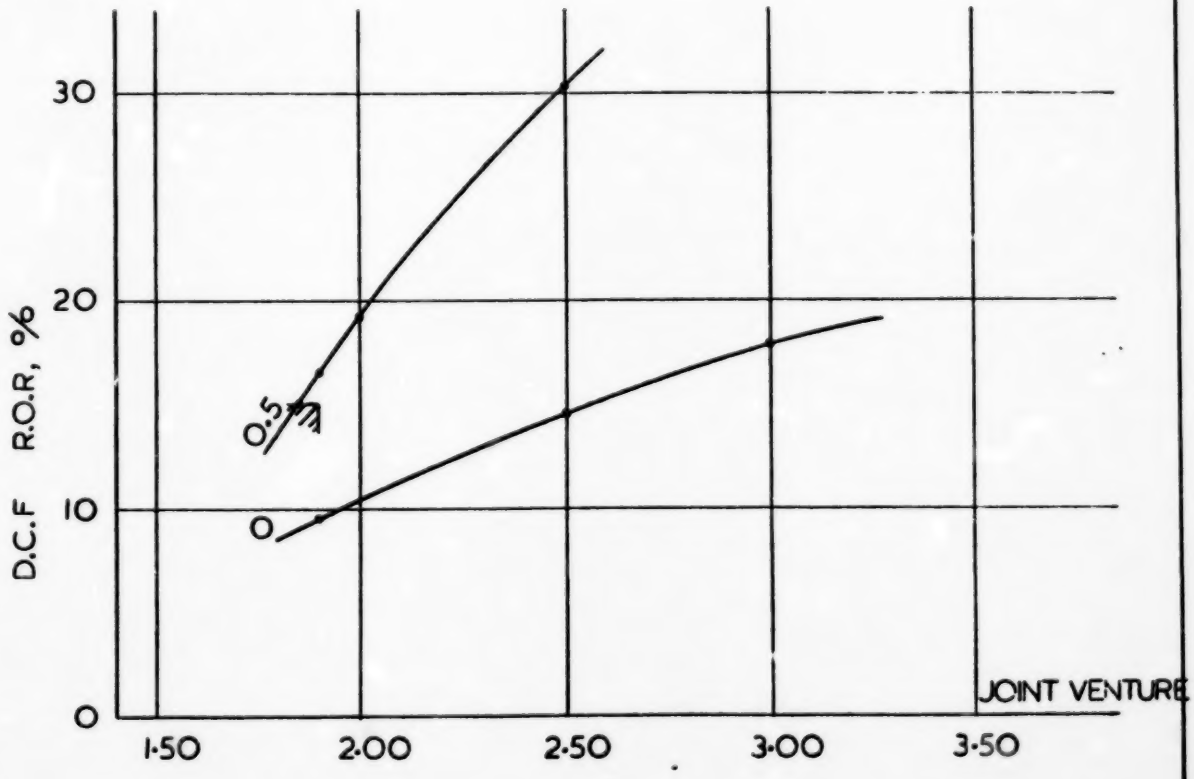
CHART-2.2.1

SNC	FEUILLE DE CALCUL	CLIENT SHELL	CONTR. 0249	SUBDIV	ELEMENT	DIV./DEPT
	DESIGN SHEET	PROJ. COAL GASIFICATION	FAIT/MADE WR M	APPR	DATE	PAGE

SUJET / SUBJECT TAX RATE=47% C.C.A=30% 0 ROYALTY DEBT/EQUITY CURVES





GAS PRICE DELIVERED ——— $/GJ

CHART-2·2·2

4000 06/0

snc	FEUILLE DE CALCUL	CLIENT SHELL	CONTR 0249	SUBDIV	ELEMENT	DIV DEPT
	DESIGN SHEET	PROJ COAL GASIFICATION	FAIT/MADE WRM	APPR	DATE	PAGE

SUJET
SUBJECT

GAS PRICE = $ 1.90/GJ
C.C.A = 50 %
JOINT VENTURE
47% TAX RATE
ROYALTY = 0
COAL VALUE = 0

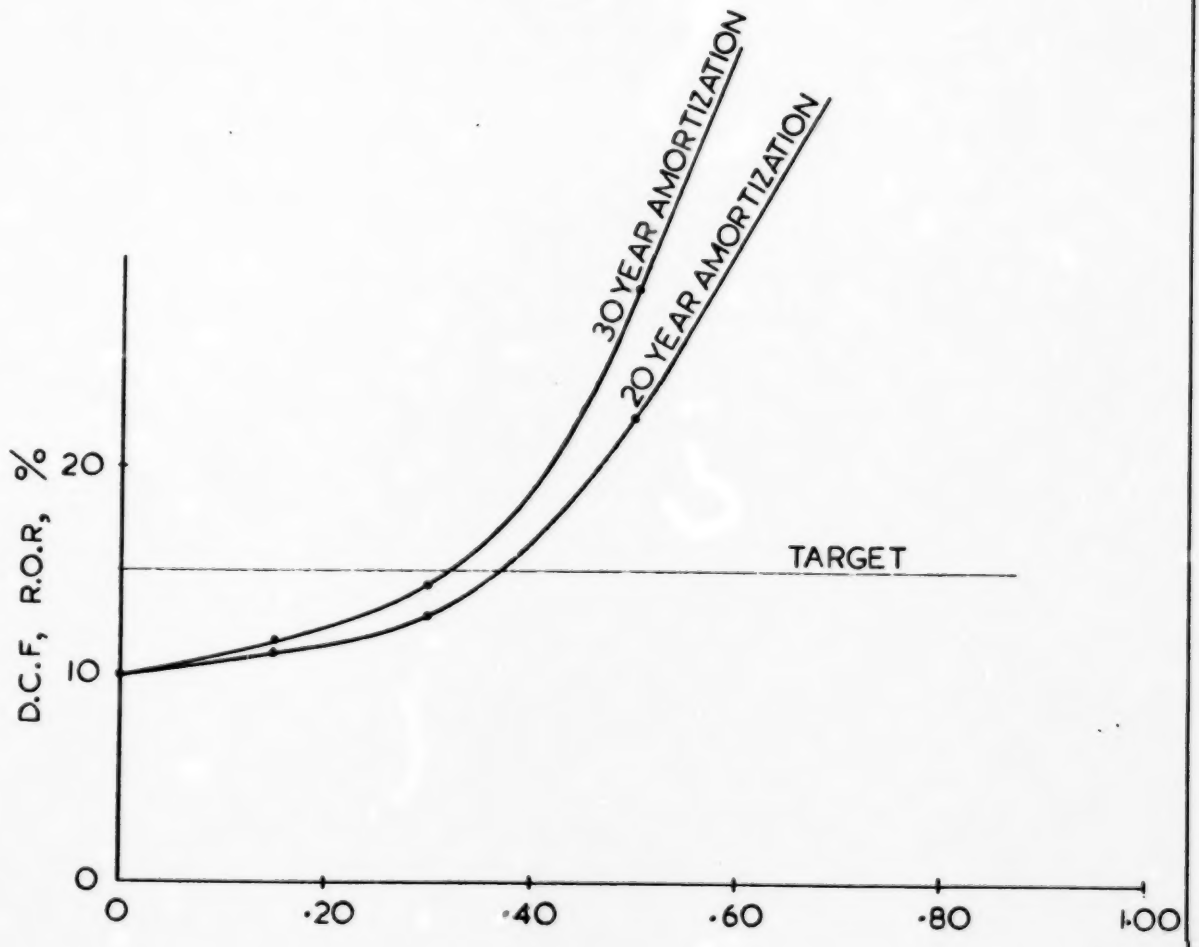


CHART-2.2.3

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	FEUILLE DE CALCUL	CLIENT SHELL	CONTR 0249	SUBDIV	ELEMENT	DIV DEPT
SNC	DESIGN SHEET	PROJ COAL GASIFICATION	FAIT/MADE WRM	APPR	DATE	PAGE

SUJET / SUBJECT: SENSITIVITIES – DEBT/EQUITY RATIO = 0

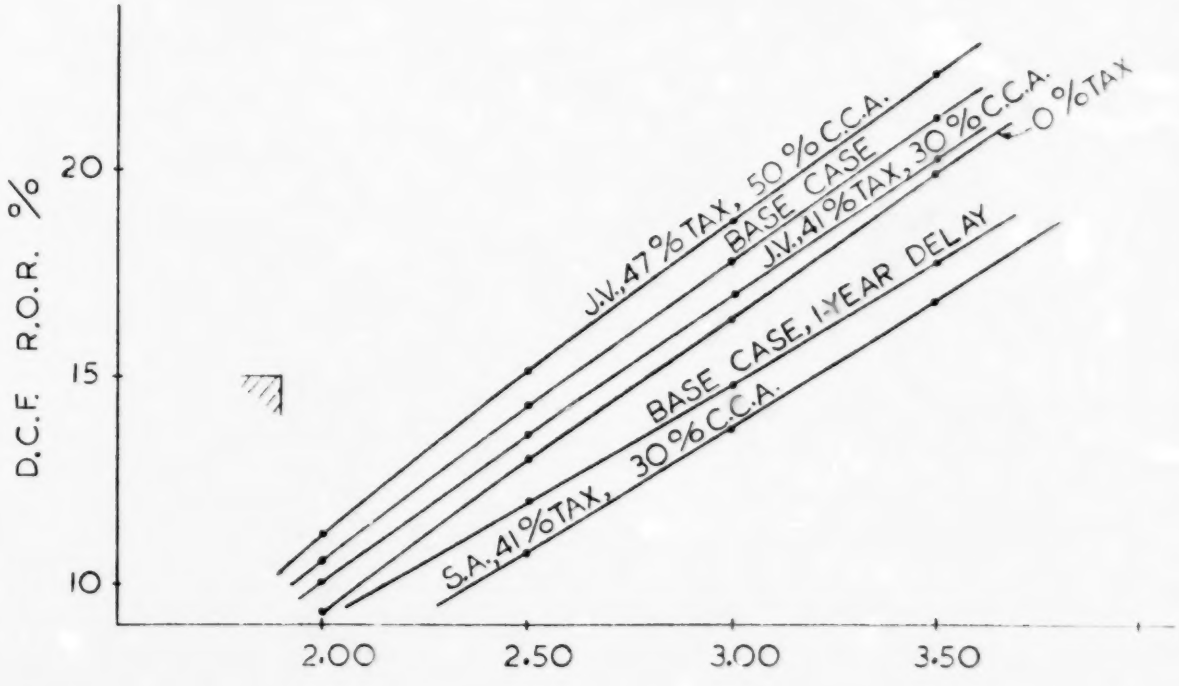

CHART-2.2.4

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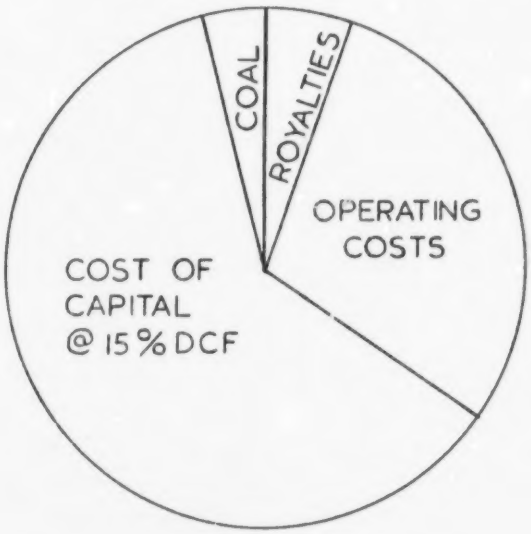
	FEUILLE DE CALCUL	CLIENT SHALL	CONTR	SUBDIV	ELEMENT	DIV./DEPT
snc	DESIGN SHEET	PROJ COAL GASIFICATION	FAIT/MADE WRM	APPR	DATE	PAGE

SUJET / SUBJECT: COST BREAK-DOWNS GAS PRICE =$2.84/GJ

BASE CASE CONDITIONS
EXCEPT COAL (IN GROUND)=$1.00
ROYALTIES ON COAL = $1.00



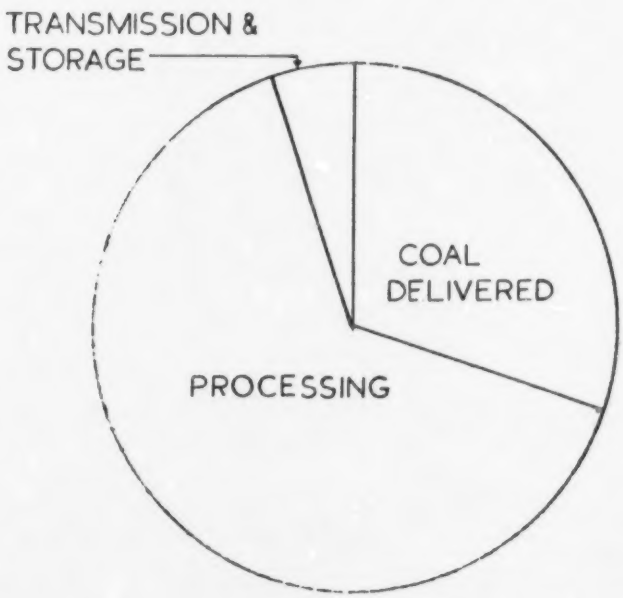


CHART-2-2-5

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3.0 CONCLUSIONS & RECOMMENDATIONS

Coal gasification using existing technology, is only marginally profitable using 1977 construction costs and energy prices. A significant reduction in capital and operating cost is necessary to meet the target D.C.F. rate-of-return under these conditions.

There are several ways in which the economics of the complex may be improved. The more promising include:

3.1 Combining the fuel gas plant with a thermal power plant.

This offers the possibility of using the fines for steam raising in the power plant, sharing the fixed mining costs and improving the overall thermal efficiency of both plants by increasing the options for recovering low grade waste heat.

3.2 Predrying the coal

Both productivity and efficiency might be improved by predrying the coal. If the moisture content of the coal were reduced to the same level as that of Sasol, in South Africa then we might reasonably expect similar productivity and efficiency. (See Table 3.1 below)

The benefits of predrying would be offset to some extent by the added capital and operating cost of the driers.

TABLE 3.1

		Edmonton	Sasol	Edmonton (dried)
1.0	Coal			
	Moisture (A.R.B.)	26.5	10.7	10.7
	Ash (A.R.B.)	17.0	32.1	20.2
	Volatiles (A.R.B.)	26.5	19.9	31.5
	H.H.V. (A.R.B.)	7050	7550	8370
2.0	Productivity (Basis 150 x 10^9 BTU/Day)			
	Number of operating gasifiers required	22	15.7	
3.0	Thermal Efficiency			
	Recovery to Gas + Liquids, percent of coal fed	82.5	89.1	

3.0 CONCLUSIONS & RECOMMENDATIONS - cont'd

3.3 Reducing the number of gasifiers

Higher productivity per gasifier can be attained either by using larger diameter gasifiers or by going to slagging gasifiers. As both of these improvements are approaching commercial reality, they should be considered in future studies.

3.4 Using a less expensive gas clean-up system

The Lurgi Rectisol system has considerable technical advantages over competing processes, but it is expensive.

3.5 Improving thermal efficiencies

Freon cycles for power generation and absorption refrigeration cycles could both be used in place of air cooling/water cooling to improve energy efficiencies. The generation of 6.9 bar steam for power generation is very inefficient and could possibly be improved in some manner.

3.6 Lower the oxygen specification

Considerable saving in capital and operating cost is possible if the oxygen specification is reduced to 95%.

3.7 Specific recommendations for future study

The primary objective of future studies should be to reduce the capital cost of the project. As improvements in technology offer good possibilities for doing this, the following is recommended:

3.7.1 Gather a large representative sample of the coal.

3.7.2 Arrange for B.G.C. to test the coal under slagging conditions.

3.7.3 Review the impact of the test results on the economics of the project.

3.7.4 Carry-out an optimization study incorporating the various improvement possibilities listed in 3.1 to 3.6 inclusive.

TABLE 4.0.2

STEAM BALANCES

.1 Level 104 bar

Production -			412.2 Kg/s
Consumption -	Extraction Turbines	393.1	
	Let-down Station	19.1	
	TOTAL	412.2	412.2 Kg/s

.2 Level 35.5 bar

Production -	Extraction Turbines	393.1	
	Let-down Station	19.1	
	TOTAL	412.2	412.2 Kg/s
Consumption	Process	393.1	
	Condensing Steam Turbines	19.1	
	TOTAL	412.2	412.2 Kg/s

.3 Level 6.9 bar

Production	Waste Heat Boilers		151.0 Kg/s
Consumption	Process	109.7	
	Condensing Steam Turbines	41.3	
	TOTAL	151.0	151.0 Kg/s

4.0 PROCESS DESCRIPTION

The overall system for supplying energy from the mine to the customer is represented by the block flow diagram, DWG. 0249-4500-11, which shows all the major streams and flowing quantities.

A preliminary layout of the plant is shown on DWG. 0249-4500-12.

The overall energy balance around the plant is shown in Table 4.0.1

TABLE 4.0.1

ENERGY BALANCE

Energy In		10^6 kJ/Hr.	
Coal	-	12,419	
Power	-	95	
TOTAL	-	12,514	
Energy Out			**Percent**
Fuel Gas	-	6,594	52.7
Tar, Oil, Naphtha	-	1,298	10.4
Ammonia	-	87	0.7
Sulphur	-	11	0.1
Phenols	-	71	0.6
Gas to Incineration	-	22	0.2
Carbon in Ash	-	101	0.8
*Export Steam	-	387	3.1
Condensers, Coolers, etc.	-	3,912	31.2
Miscellaneous	-	31	0.2
		12,514	100.0

*This steam is passed through a condensing turbine to produce 15MW of power used in the plant.

A brief description of the various parts of the system follows.

An overall material balance is not included. The approximate disposition of materials is shown on Drawing 0249-4500-11. An accurate material balance would require testing of the coal by the gasification process licensor.

A steam balance is shown in table 4.0.2 and a water balance in table 4.0.3.

TABLE 4.0.3

WATER BALANCE

WATER IN

Make-up from river	227.9	Kg/s
In coal	42.9	Kg/s
	270.8	Kg/s

WATER OUT

Decomposed	100.7	Kg/s
Cooling tower evaporation	87.3	Kg/s
With ash	9.0	Kg/s
To incineration	2.0	Kg/s
With ammonia	3.3	Kg/s
With tar	0.2	Kg/s
To disposal well	47.5	Kg/s
Process evaporation	4.1	Kg/s
Pond evaporation	8.9	Kg/s
Unaccountable	7.8	Kg/s
	270.8	Kg/s

* Values are approximate and subject to review by Lurgi.

4.0 PROCESS DESCRIPTION - cont'd

4.1 Mining

4.1.1 General

The complete mining plan was produced by the coal mining consultants, H.G. Stephenson and W.R. Hardcastle. As this work had to be carried out in parallel with other portions of the project, a mining rate had to be assumed which subsequently turned out to be in error. Consequently, the reserves included in this report will only support the plant for 25 years. As there is more developable coal in the area this variation from the original basis is not considered to be serious.

Additions to the capital and operating cost have been made to handle the additional tonnage.

4.1.2 Mining Rate

Approximately 18,000 metric tons per day or 5,998,000 metric tons per annum of coal is required to produce the desired quantity of fuel gas. With 200,000,000 short tons of minable coal included in the reserves, the mine as outlined will feed the plant for over 25 years on an 85% recovery basis.

4.1.3 Mining Method

It is proposed to operate the mine as two separate open pits. The physical development of the reserves, the significantly different ash contents in each of the mining areas, and the flexibility that two mines provide all support this method.

A combination of large 58 cubic yard electric walking draglines, 13 cubic yard electric walking draglines, 14 cubic yard electric shovels plus front end loaders and bulldozers make up the mining equipment. 100 Ton bottom dump coal haulers will be used to haul the coal to the plant site and 75 ton rear dump haulers will be used to return the ash to the mine.

The large draglines will be used mainly for stripping, and the other equipment for mining the coal and removing partings.

The stripping ratio varies from a low of 4.5 to 1 to a high of 15 to 1, with the average for both sites 8.5 to 1.

TABLE 4.1.1

COAL QUALITY

	POOREST SITE (wt. %)	AVERAGE (wt. %)
Moisture	26.5	26.5
Ash	21.0	17.0
Volatiles	24.6	26.21
Fixed Carbon	27.9	30.0
Sulphur	0.29	0.29
Calorific Value (BTU/lb.)	6500	7050

TABLE 4.1.2

MINERAL ANALYSIS OF ASH

		(wt. %)
Silicon Dioxide	-	51.45
Aluminum Oxide	-	19.93
Iron Oxide	-	5.26
Calcium Oxide	-	7.51
Sodium Oxide	-	3.67
Sulphur Trioxide	-	5.61
Magnesium Oxide	-	1.04
Potassium Oxide	-	0.70
Titanium Oxide	-	0.60
Phosphorus Pentoxide	-	0.20
Manganese Oxide	-	0.02
Chromium Oxide	less than	0.01

4.0 PROCESS DESCRIPTION - cont'd

4.1 Mining - cont'd

4.1.3 Mining Method - cont'd

The reclamation plan is to fill the low areas with ash, cover the ash with spoil, level the spoil, and cover the spoil with reserved top soil. The land can then be sown with grass or otherwise cultivated.

4.1.4 Coal Quality

There is considerable variation in the ash content of the coals from one site to the other. By mining the two sites simultaneously, it will be possible to blend the two coals thus minimizing the variations at the plant. The coal quality on an as-received basis is shown in Table 4.1.1.

The operating cost of coal delivered to the plant has been estimated at $2.60 per metric ton exclusive of royalties, taxes, interest, return on investment or depreciation.

4.1.5 Coal Preparation

Coal from the mine is dumped into a pocket and stockpiled by a stacker. Approximately two days supply can be kept in the stockpile.

Reclaimed coal is conveyed to a screening tower where the plus 50 mm coal is scalped, crushed and returned to the screens. The minus 50 mm plus 3 mm coal is fed to the gasifiers. The minus 3 mm coal is conveyed to the powerhouse where it is used as fuel for the two high pressure boilers. Approximately 24% of the coal fed to the screens will be fines and will be burned in the boilers.

4.2 Coal Processing

4.2.1 Coal Gasification and Gas Purification

The sized coal is fed into the Lurgi pressurized fixed bed gasifiers through lock hoppers where it is gasified by reacting with steam and oxygen. The coal moves counter to the flow of gases so that combustion occurs at the bottom of the reactor, gasification higher up and devolatilization and drying at the top of the reactor. Ash is quenched with water and removed through a lock hopper at the bottom of the reactor.

4.2 Coal Processing - cont'd

4.2.1 Coal Gasification and Gas Purification - cont'd

The resultant gas contains carbon monoxide, carbon dioxide, hydrogen, methane, ethylene, naphtha, oils, tar and some ash. It is first quenched with an aqueous liquor to condense the tars and scrub both the tars and the ash. The quenched gas is further cooled in a waste heat boiler producing 6.9 bar steam, then in an air cooler and finally in a water cooler.

Sufficient quench liquor is bled from the system to maintain a balance. It is cooled, and flashed to atmospheric pressure. The dissolved gases released are scrubbed to recover ammonia, then incinerated. A physical separation is made between the heavier-than-water tars and solids in conical separators, and the lighter-than-water oils in corrugated plate intercepters and coalescers. The tars and oils go to storage for sale while the waters go to the Lurgi Phenolsolvan process for further cleanup.

The aqueous stream first passes through a gravel bed filter and then enters a counter-current solvent extraction unit which removes phenols from the waters. The dephenolated water is stripped with nitrogen for solvent recovery and sent to the waste water treatment plant.

Condensate from the gas cooling train is similarily treated but in addition, it is stripped of H_2S in a deacidifier, then of ammonia in an ammonia stripper, and finally of dissolved hydrogen, carbon monoxide and methane in a spray cooler.

The two streams are treated separately because the solids in the quench liquor stream greatly inhibit biological action in waste water treatment.

The stripped H_2S goes to sulphur recovery and the aqueous ammonia goes to the ammonium sulphate plant.

The cooled gas is next treated in the Rectisol plant which recovers naphtha and removes carbon dioxide and H_2S.

The crude gas is chilled then scrubbed with cooled methanol to remove water and naphtha. Naphtha and methanol are separated in an azeotrope column, and water and methanol in a distillation column.

4.2 Coal Processing - cont'd

4.2.1 Coal Gasification and Gas Purification - cont'd

The naphtha-free gas is chilled further with colder methanol to absorb H_2S and CO_2. The acid gases are stripped from the methanol and sent to the Stretford unit for sulphur recovery. The sulpur-free gas enters the distribution pipeline at 28.6 bar (400 psig).

In the Stretford process, the H_2S is absorbed by direct contact in an aqueous alkaline solution. The H_2S is oxidized to elemental sulphur, the sulphur is separated from the aqueous phase and sent to the sulphuric acid plant, and the alkaline solution is regenerated for reuse.

With 99.5% of the H_2S removed, the carbon dioxide stream containing a small amount of hydrocarbons is incinerated.

A considerable quantity of gas is vented from the lock hoppers when they are depressured. These gases are scrubbed in a Venturi scrubber to remove solids, then compressed for recycle. A low pressure gas holder is necessary to smooth out surges.

The gas is first compressed to 14.8 bar then mixed with $0^{o}C$ gas from the Rectisol unit and the mixed gas is compressed to 32.5 bar.

4.3 Auxiliary Plants

The gas producing plant requires several other auxiliary units. Steam, oxygen, power, cooling water, boiler feed water, refrigeration and waste water treatment are all required.

4.3.1 Oxygen Plants

Two large air separation plants provide the approximately 2600 metric tons per day of oxygen consumed in the gasification plant as well as the smaller quantity of nitrogen required.

4.0 PROCESS DESCRIPTION - cont'd

4.3 Auxiliary Plants - cont'd

4.3.2 Power Plants

Two large boilers producing a total of 742,000 kg/Hr. of steam at 104 bar and 510°C are required. These boilers are designed to be fired with pulverized coal and include the necessary de-aerating, pumping, electrostatic precipitating, ash disposal and other anciliary equipment.

High pressure super-heated steam is passed through two back pressure turbines exhausting at approximately 35 bar to supply the required saturated process steam for the gasifiers and at the same time generate 40 MW of power.

Excess medium pressure steam produced in the gasification section is exhausted through a condensing turbine to produce an additional 15 MW of power.

A water treatment plant designed to handle water from the North Saskatchewan River is also included.

4.3.3 Ammonium Sulphate Plant

The aqueous ammonia produced by the Phenolsolvan plant can be treated in several ways: concentrated by distillation to anhydrous ammonia, reacted with sulphuric acid to produce ammonium sulphate, or reacted with phosphoric acid to produce ammonium phosphate. Production of ammonium sulphate was chosen because sulphur was available and inexpensive, the salt has a ready market, storage as a solid is simple and impurities in the aqua are easily handled.

The plant consists of a reactor for the aqua and sulphuric acid and a triple effect evaporator-crystalizer combination to produce crystals. The crystals are separated from the magma by DSM screens and centrifuges, dried, coated and transferred to storage. A description of the process and a flowsheet is included in the Appendix.

4.3.4 Sulphuric Acid Plant

A double absorption sulphuric acid plant capable of providing about 280 metric tons per day of 66 Be sulphuric acid is required to neutralize the recovered ammonia. The plant will consume all the sulphur produced from the Stretford unit plus approximately an equal amount of purchased sulphur.

4.0 PROCESS DESCRIPTION - cont'd

4.3 Auxiliary Plants

4.3.5 Cooling Water System

A cooling water system is required for the gasification plant. Approximately 20,000 IGPM of cooling water is required with a temperature drop of 29°F across the tower.

4.3.6 Refrigeration System

Two levels of refrigeration are required for the Rectisol plant - one at 0°C and one at -42°C. Approximately 3,000 HP is required to drive the refrigeration compressor.

4.3.7 Waste Water Treatment

The dephenolated quench water is first treated in a froth flotation plant to remove residual oils, tars and solids, before joining the rest of the system waste water in a two-stage aerated activated sludge biological treatment plant.

Waters not suitable for recycling are rejected down a disposal well.

4.3.8 Pipelines

A gas pipeline network connecting the gasification plant to customers in the Fort Saskatchewan, Redwater and Edmonton areas has been sized to consume the available pressure drop. The pipelines will follow existing corridors. The line sizes vary from 36" in diameter at the plant site to 16" in diameter for one supply header.

Included in the cost estimate is the cost of moving existing pipelines in the mining area.

A 16" water pipeline complete with river water intake and a pumping station is also required to supply the site with water from the North Saskatchewan River.

4.3.9 Gas Storage

In order to decouple the gasification plant from the consuming plants, salt cavern storage will be used. Gas will be compressed to 220 bar. Capacity of the cavern will equal six weeks production at half rate. When recovering gas from the caverns, it will be dried before entering the distribution system.

5.0 OPTIONS

5.1 Site Selection

Although it seemed likely that a mine mouth site would be optimum, the possibility of siting the plant at Fort Saskatchewan close to the major consumers was considered.

Two methods of transporting the coal were examined:

1. Slurry pipeline
2. Unit train

Slurry pipelines were non-competitive with unit trains because of the considerable expense required to dewater the coal. This expense becomes progressively less important as the distance increases, so that pipelines appear to be the preferred method of moving large volumes long distances.

The unit train concept, although better than slurry pipelining is not competitive with pipelining the gas to the plant and water to the site. The best price the railroads would give was between $5.00 and $5.50 per short ton of coal, which would exceed the pipeline capital cost in the first year of full operation.

The possibility of operating a separate dedicated railroad was also explored and rejected as it would be almost as expensive in capital as a pipeline, it would have a higher operating cost, it would be opposed by the existing railroad companies, and it would not remove the need to pipeline gas to both Redwater and Edmonton.

5.2 Gasifier Selection

Gasification may be carried out in a fixed bed, an entrained bed, or a fluidized bed. All these methods have been proven in commercial operation.

In the fixed bed system as developed by Lurgi, pea-sized coal is fed into the top of the reactor, where it slowly descends, contacting an upward flow of hot gas and steam. There are three zones in the reactor: an upper zone where the coal is preheated and devolatilized, a middle zone where the coal is gasified, and a bottom zone where residual carbon is combusted to provide the necessary heat to carry out the reaction.

In the entrained bed system as developed by Koppers-Totzek, finely pulverized coal is entrained in a gasifying stream of steam and oxygen. The mixture is blown into a refractory lined gasifier where a sequence of oxidation and steam decomposition reactions occurs. Effective operation requires high temperatures, normally above the ash fusion point.

5.2 Gasifier Selection cont'd

In the fluidized bed system such as the Winkler, a bed of finely divided coal is fluidized by a gasifying medium consisting of steam and air or oxygen which flows upwards through it with a controlled velocity. The bed is continuously subjected to rapid agitation and mixing which results in high heat and mass transfer rates. Temperatures are essentially constant throughout the bed, and must be kept low enough to prevent excessive ash agglomeration.

All three systems have both advantages and disadvantages.

The Lurgi system uses the least oxygen but the most steam, and requires the least coal preparation. Operating at higher pressures and lower temperatures, it produces a gas richer in methane and therefore heating value. High ash coals are handled without difficulty, but caking coals require pretreatment. There are many large scale units in commercial operation. The lower operating temperatures also result in the production of tars, phenols and fatty acids, however, which are difficult to recover and cause many disposal problems.

The K-T system uses the most oxygen and the least steam, and requires the most elaborate coal preparation system as the coal must be both pulverized and dried. The system operates above the ash melting point, with each particle separated from the rest, so it can handle caking coals. High ash coals are troublesome, however, as significant quantities of ash carry-over to the waste heat boilers and cause erosion problems. The gasifiers operate at essentially atmospheric pressure. As with the Lurgi system, there are many large scale units in commercial operation. The high temperature operation results in total gasification, therefore eliminating the problems associated with tars and phenols, but producing a gas with essentially no methane content and containing some Hydrogen Cyanide which must be accommodated in some way.

The Fluidized Bed System operates at conditions between the other two. It requires intermediate quantities of oxygen and steam, operates at reaction rates that are faster than Lurgi but slower than K-T. It has the potential for developing extremely large size single trains following the example of Fluid Catalytic Cracking and Fluid Coking technology. Like K-T, it can be operated to produce few tars, but can be richer in methane, though not as rich as Lurgi. There are few large scale production units in operation, however. A large portion of the ash content is carried out of the reactor with the gas, and a significant portion of the carbon in the feed must be removed with the ash in order to keep carbon concentrations high enough to allow the steam decomposition reaction to take place. High ash coals therefore cause high carbon losses. Caking coals cannot be handled without pretreatment and the gasifier, like K-T, operates at essentially atmospheric pressure.

The need for a high pressure gas transported approximately 50 kilometers to market with a heating value sufficiently high to be used in natural gas fueled gas turbines, produced from a non-caking coal high in both ash and moisture, tilted the balance in favor of the Lurgi fixed bed process.

5.3 GAS CLEANING

Several methods of cleaning the gas were considered before the Rectisol cold methanol wash - Stretford sulphur recovery systems were chosen.

Undesirable materials in the gas include Hydrogen Sulphide, ammonia, tars, oils, phenols, and other condensibles. After quenching and cooling, the materials remaining are mainly Hydrogen Sulphide, and some light naphtha. In addition, in order for the fuel gas to reach the required heating value, some but not all of the Carbon Dioxide present must be removed, and the gas must be dried before entering the pipeline.

The Alkazid process was initially selected. The process operates in an absorption - stripping cycle similar to the familiar amine processes for acid gas removal, but employs different solutions. Solution "DIK.", which contains the potassium salts of diethyl or dimethyl glycine, can be used to selectively remove Hydrogen Sulphide in the presence of Carbon Dioxide, Carbon Disulphide or Hydrogen Cyanide. The process is simple, with a relatively low capital cost, but is reported to have severe corrosion problems. The process was discarded because it could not produce a rich enough Hydrogen Sulphide stream for a Claus sulphur plant, because of the low sulphur level in the coal.

The Stretford process uses an alkaline solution of anthraquinone disulphonic acid and sodium metavanadate to scrub Hydrogen Sulphide from the gas stream and oxidize it to elemental sulphur. In this process, vanadium is reduced from the five valent state to the four valent state, and sulphur is precipitated as a finely divided solid. The sulphur is removed from the solution, washed, melted, and recovered for sale or further processing. The solution is regenerated in a column by oxidation with air before recycling.

The Stretford process has as it's chief advantage the ability to remove virtually all of the Hydrogen Sulphide present in even very dilute gas streams, and convert it directly to elemental sulphur. Although it is not normally used to remove Carbon Dioxide, it reportedly can do so to some extent. Unfortunately, this possibility could not be confirmed within the budget of this study.

The Rectisol process was ultimately chosen to clean the gas because it could condense the naphtha, remove water vapour and absorb both the Carbon Dioxide and Hydrogen Sulphide in a system using cold methanol as the single absorbent, because the system is very reliable with few operating difficulties, and because the design and operating data for the process in this service is well known.

As the acid gas produced is too dilute to feed a Claus sulphur plant, the Stretford process was chosen to scrub and convert the Hydrogen Sulphide from this gas prior to incineration.

In future studies it may be beneficial to review this decision when more information has been received on the Stretford process.

5.4 Power Balance Options

After due consideration of the several options open, it was concluded that the optimum power configuration was to use the fines from the screening process to raise steam at 1500 psig, to produce power by exhausting the steam through a back pressure turbine to 500 psig, to use the excess 100 psig steam from the waste heat boilers in the process to produce more power by passing it through a condensing turbine, and to purchase the remaining power required.

The possibilities considered in arriving at this conclusion were the following:

1. Burn all the coal fines in the boiler, sell the liquids at their fuel value, use the excess low pressure steam to produce the maximum power possible, and purchase the balance required.

2. Burn the liquids in a combined cycle gas turbine, burn sufficient fines to produce the necessary steam in the boilers, produce all the power possible with low pressure steam, and sell the balance of the fines remaining as well as the excess power produced.

3. Burn the liquids in a combined cycle gas turbine, burn all the fines in a boiler, and sell the excess power produced.

4. Burn the liquids in a combined cycle gas turbine, burn the fines required to produce the deficiency in steam, gasify the remainder in a fines gasifier and sell the excess power produced.

5.5 By-Products

Tar, oil, naphtha, crude phenols, ammonia and sulphur are co-produced with the fuel gas. These materials are similar to those produced from coke oven gas and therefore have significant economic potential. However, as they are produced at temperatures very much lower than in coke ovens, they are different enough to present a serious marketing problem.

5.0 OPTIONS - cont'd

5.5 By-Products - cont'd

5.5.1 Tar

The tar is a heavier-than-water organic stream produced by decanting quench liquor and is probably the least valuable of the by-products produced, mainly because of the solids content (coal dust and ash). Traditionally, coal tar has been used for the production of creosote, as a road tar prime, and as a fuel.

Creosote is probably the best constituent in the tar but as its quality differs significantly from that presently in use, a major marketing effort would be necessary to gain acceptance of the material.

Other tar constituents can be used in the production of electrode pitch and carbon black, but the most likely disposition of this material is as a fuel.

5.5.2 Oil

The oil is a lighter-than-water organic stream that condenses in the gas cooling train. It contains significant concentrations of benzene and naphthalene precursers which could be produced at high quality if the material were used as hydrodealkylation feed to a benzene unit.

It should be possible to market this oil at a crude oil price.

5.5.3 Naphtha

Naphtha is a light organic stream containing olefins, paraffins, and aromatics, recovered from the fuel gas by a low temperature methanol wash in the Rectisol unit.

This material is analogous to condensate recovered from natural gas processing plants, and should command a similar premium. It can be used as feed to a benzene plant, feed to an olefins plant, as a diluent or separation fluid for transporting other petroleum liquids, or as a component in motor fuel.

5.0 OPTIONS - cont'd

5.5 By-Products - cont'd

5.5.4 Crude Phenols

The crude phenols stream is a heavier than water organic material consisting of tar acids, tar bases, (pyridines, picolines) neutral oils, organic acids and pitch recovered from the gasification plant aqueous liquors by solvent extraction in the Phenolsolvan unit.

This stream could be processed to produce neutral oil - free phenol and refined cresylics for sale, but would most probably be sold as-is to an existing processor currently using coke oven and/or petroleum by-product phenols.

5.5.5 Ammonia

Ammonia is produced from the Phenolsolvan process as a 23% aqua. This material has no appreciable local market, but can easily be upgraded to ammonium sulphate by reacting with sulphuric acid.

5.5.6 Sulphur

Elemental sulphur is produced from the Stretford process. This sulphur would be essentially unsaleable in Alberta, because of a severe over supply in the province. It is proposed to dispose of it by converting it to sulphuric acid and combining it with aqua to produce ammonium sulphate.

5.5.7 Ash

Ash is produced both from the gasifiers and from the boilers. It can be used to produce bricks, as a road building base, as an extender for putty, as a filler in either rubber manufacture or for fertilizers, as a constituent of cement, or as a constituent of metallurgical flux.

No substantial revenue is expected from the ash. Some may be used for building mine roads, while the balance will become landfill for the mined-out area.

6.0 ENVIRONMENTAL IMPACT

The coal processing plant will need to satisfy a number of environmental restrictions including, land reclamation of the mined-out area, ground water contamination and/or diversion, atmospheric emissions including fugitive dusts, liquid emissions, and river water usage.

6.1 Land Reclamation

Land reclamation and ash disposal must be considered together. A reclamation procedure similar to other prairie strip mines has been assumed.

Ash will be dumped progressively to fill a low area off the coal field near the plant. It will be covered with a layer of overburden from the mine, then by topsoil.

In the mine area, spoil piles will be contoured to be suitable for farm tractors, then covered with reserved topsoil and seeded to grasses.

Part of the cost of reclamation will be the posting of a performance bond with the Alberta Government to insure that satisfactory reclamation standards are maintained.

6.2 Groundwater

A study will be necessary to insure that the mine does not adversely affect the quality and supply of groundwater that is presently used by farmers and communities in the area.

As aquifers generally follow coal seams, mining will undoubtedly disturb groundwater conditions. However, the disturbance is unlikely to extend any appreciable distance from the mine.

Of potentially more serious concern is the possibility of contaminating the ground with phenols, pyridines, picolines, etc. when disposing of the ash. Although there is some evidence that in fact ash absorbs some of the contaminents in the water, for this study, only treated water is assumed to be used for quenching the ash.

A study of the hydrology of the area (costing between $50,000 and $150,000) would be required as part of the approvals program.

6.3 Atmospheric Emissions

The major atmospheric emission from the plant will be flue gas from the boilers. These gases will contain sulphur dioxide, but well within the limits set by the Department of the Environment. Fly ash will be recovered by electrostatic precipitators.

6.0 ENVIRONMENTAL IMPACT - cont'd

6.3 Atmospheric Emissions - cont'd

Some testing may be necessary to insure that the ESP's will perform adequately on the fly ash.

Odour control will be an environmental problem, with the main offender being the process waters. They will have to be contained until after the major odoriferous materials have been removed.

Coal dusts will have to be contained. The conveyors, screening and crushing areas will have to be covered. Siting of the mine roads and stockpile area will reflect the need to minimize the generation of fugitive dusts.

6.4 Liquid Emissions

The remoteness of the site from a large water source encourages the re-use of water. Consequently, all streams will be treated for re-use except those containing large quantities of dissolved solids, e.g. cooling tower blowdown, which will be pumped into a disposal well.

6.5 Water Supply

The water supply is assumed to be the North Saskatchewan River. At this point in time, the river has a more than adequate supply to meet all current demands plus the anticipated needs of the coal gasification plant.

However, a hydrology study will be necessary and public hearings held before approval to divert water will be given. If heavy industrial demand is established before a decision is made to build, it is conceivable that approval may be denied, in which case the line would have to come from the Pembina River.

7.0 THE EFFECT OF INTERMEDIATE BTU GAS ON THE CONSUMER

7.1 General

The primary use of the manufactured gas would be as a fuel with the possibility of using it also as a chemical feed stock for natural gas based processes. Considering the gas primarily as a fuel removes the need to process it to SNG quality, which improves thermal efficiency and reduces costs. However, the resultant gas is significantly different from natural gas (see Table 7.0.0), which will force changes by the consuming plants.

7.2 Natural Gas vs. Fuel Gas

The major differences are in heat contents of the gases (424 BTU/SCF vs 1009 BTU/SCF) and in the specific gravity of the gases (0.4 vs 0.6).

The fuel gas has some advantages over natural gas. For a given heat release the stoichiometric combustion air requirement is less, and the theoretical flame temperature is higher. Both of these differences tend to increase the capacity of a given boiler or process heater. The lower air requirement reduces the flow of flue gas, while the higher temperatures increase the rate of heat transfer.

The lower heat content, however, is a disadvantage in delivering the gas to the boiler or heater.

7.3 Distribution System

The main problem facing the consumer when switching from natural gas will be to modify the distribution system to handle the increased flow required.

Either bigger lines and control valves or higher pressure drops will be necessary. Each consumer will have to check his system, but as the gas will be supplied to the property line at 300 psig, most consumers will probably be able to use the existing distribution system with higher pressure drops.

If the gas is to be used in a gas turbine, however, it must be delivered at a higher pressure, which will likely require bigger pipe lines.

TABLE 7.0.0

GAS COMPARISONS

PROPERTY	NATURAL GAS (Typical)	FUEL GAS
H.H.V., Btu/SCF	1009	424
S.G. (air = 1.0)	0.6	0.4
COMPOSITION	**VOLUME %**	**VOLUME %**
H_2	Ø	58.05
CO	Ø	23.04
CO_2	0.04	2.94
C_1	92.24	15.18
C_2	2.77	0.42
C_3	0.8	Ø
C_4	0.28	Ø
C_{5+}	0.09	Ø
N_2	3.78	0.36
Theoretical Air Required to Release 1,000,000 Btu (SCF)	9,790	8,220
Theoretical Flame Temp, ^{o}F	3,550	3,900

7.4 Burners

Some of the combustion properties of the components in natural gas and intermediate BTU gas are shown in Table 7.4.0.

Ignition temperature may be defined as the lowest temperature at which heat is generated by combustion faster than heat is lost to the surroundings, and combustion becomes self-propagating. This temperature is affected by a large number of factors, so that the values shown are not necessarily valid except under the specific conditions of the test under which they were determined.

However, ignition temperatures do provide a measure of the tendency of hot objects to ignite a gas, and therefore are important in comparing gases as fuels. In this case, the values are close enough to ignore ignition temperatures when comparing manufactured fuel gas with natural gas.

The term rate of flame propagation, ignition velocity and flame speed all refer to burning velocity, which is dependent upon the gas-to-air ratio, the size and shape of the container, the test method employed, the temperature of the mixture, and energy-absorbing sinks near the flame.

Each of the combustible constituents in a fuel gas has a rather definite burning speed under given conditions. From Table 7.4.0, it can be seen that hydrogen is by far the fastest burning gas, followed by carbon monoxide. It can therefore be concluded that the manufactured fuel gas will burn at a higher velocity than will natural gas.

This higher velocity affects the burner design and also the heating characteristics of the flames. The main effect is to move the position of highest temperature closer to the burner, which is generally a favourable change. Before accepting this conclusion, however, each furnace should be examined for sensitivity to shorter flames.

Specific flame intensity characterizes the concentration of heat available from a flame. As with burning velocity, hydrogen has by far the highest rate of heat release of the constituents present in both natural gas and the fuel gas. The manufactured fuel gas will therefore have a more compact flame than would natural gas.

Burners that use gas velocity to induce combustion air will aspirate far too much air when switched to manufactured mainly because of the high gas flow, but also because of the lower air requirements. This type of burner will need adjustment and possibly replacement of orifices.

All the burners will need review to ensure good operation using manufactured fuel gas. Most, however, should be able to use the lower heating value fuel with minor adjustments or modifications.

TABLE 7.4.0

COMBUSTION CHARACTERISTICS

COMPONENT	IGNITION TEMP., ^{o}F (in Air)	RATE OF FLAME PROPAGATION, f.p.s. (Burner Method)	SPECIFIC FLAME INTENSITY (MAX. BURNING VELOCITY MIXTURE) BTU/SEC - ft^2
H_2	1060	9.2	703
CO	1191 - 1216	1.7	124
C_1	1301	0.85	126
C_2	968 - 1166	1.06	143
C_3	871	0.95	145
C_4	864	1.05	144

7.0 THE EFFECT OF INTERMEDIATE BTU GAS ON THE CONSUMER - cont'd

7.5 Gas Turbines

GE gas turbines will operate successfully with fuel gas with heating values as low as 300 BTU/SCF. The main difficulty, therefore, will be to supply the combustor with sufficient gases which will probably mean re-piping the machines with larger lines and control valves.

7.6 Process Gas

The manufactured gas is already partially converted to synthesis gas, so it should be a good feed stock for hydrogen, ammonia or methanol production. The fuel requirement in a reformer furnace should be significantly lower than for natural gas feed stock.

However, the gas contains more sulphur than natural gas, which must be removed.

A catalyst bed containing a cobalt-molybdenum catalyst to convert all sulphur present to hydrogen sulphide followed by a zinc oxide bed to absorb the H_2S would probably have to be added to any synthesis gas plant before converting from natural gas.

SIZE "D" 22" x 34"

GENERAL NOTES & SPECIFICATIONS

POWER 96.0 MW
EVAP
OXYGEN PLANT
NITROGEN 9.5 kg/s
OXYGEN 80.22 kg/s
TO DISPOSAL WELL
9.0 kg/s
POWER 0.3 MW
CRUSH
217 Mg/d
MINE 18176 Mg/d
HAUL
STOCK PILE 30,000 TONNES
RECLAIM
SCREEN
15,996 Mg/d
4130 Mg/d
POWER 9.6 MW
GASIFICATION QUENCH & COOL
207.76 kg/s
ASH
GAS
6.9 BAR STEAM
4.13 kg/s
MISCELLANEOUS STEAM USERS
COND.
680 kg/s
POWER 2.4 MW
REFRIGERATION
9.99 kg/s LIQUID NH3
5.82 kg/s NH3 VAP (-29°C)
3.73 kg/s NH3 VAP (0°C)
INCINERATION
SULPHUR RECOVERY
SOUR GAS
NAPHTHA 125.99 Mg/d
2.18 kg/s
GAS (21 BAR)
199 x 10⁹ kJ/d
POWER 7.9 MW
STORAGE 220 BAR
COMPRESSION
DRYER
149.37 x 10⁹ kJ/d
4,990.89 Mg/d
GAS MARKET
SULPHUR 46.2 Mg/d
SULPHUR STORAGE
47.92 Mg/d
SULPHUR
SULPHURIC ACID PLANT
STEAM
2778 Mg/d ACID
93.97 Mg/d
CRUDE PHENOLS
YARMOUTH
0.99 kg/s
STORAGE 75,000 TONNES
AMMONIUM SULPHATE PLT
AMMONIUM SULPHATE
379.84 Mg/d
YARMOUTH
AQUA
CONTAMINATED COND.
CONDENSATE
TAR
OIL
WATER 108.87 kg/s
LIQUIDS BY-PRODUCTS RECOVERY
30.6 kg/s
FROTH FLOTATION
20.7 kg/s WATER
1.5 kg/s
0.1 kg/s
MINE 3090 Mg/d(DRY)
HAUL
STORE
EVAP
POWER 0.8 MW
COOLING TOWER
1524 kg/s CWS
1524 kg/s CWR
46.6 kg/s
7.5 kg/s BLOWDOWN
BIO-OXIDATION
POWER
INTAKE
RIVER PUMP HOUSE
227.88 kg/s
WATER RESERVOIR 900,000 m³
99.99 kg/s
208 kg/s
PUMP HOUSE
104.28 kg/s
0.99 kg/s
18.6 kg/s
315.99 kg/s
WATER TREATMENT
216.6 kg/s
POWER PLANT
34.5 BAR STEAM
POWER 35 MW
TO PLANT
PURCHASED 24 MW
DISPOSAL WELL
20 kg/s
10.92 kg/s
TAR 8,000 m³
OIL 8,000 m³
NAPHTHA 8,000 m³
STORAGE TANKS
329.20 Mg/d
182.34 Mg/d
125.99 Mg/d
MARKET
MARKET
MARKET

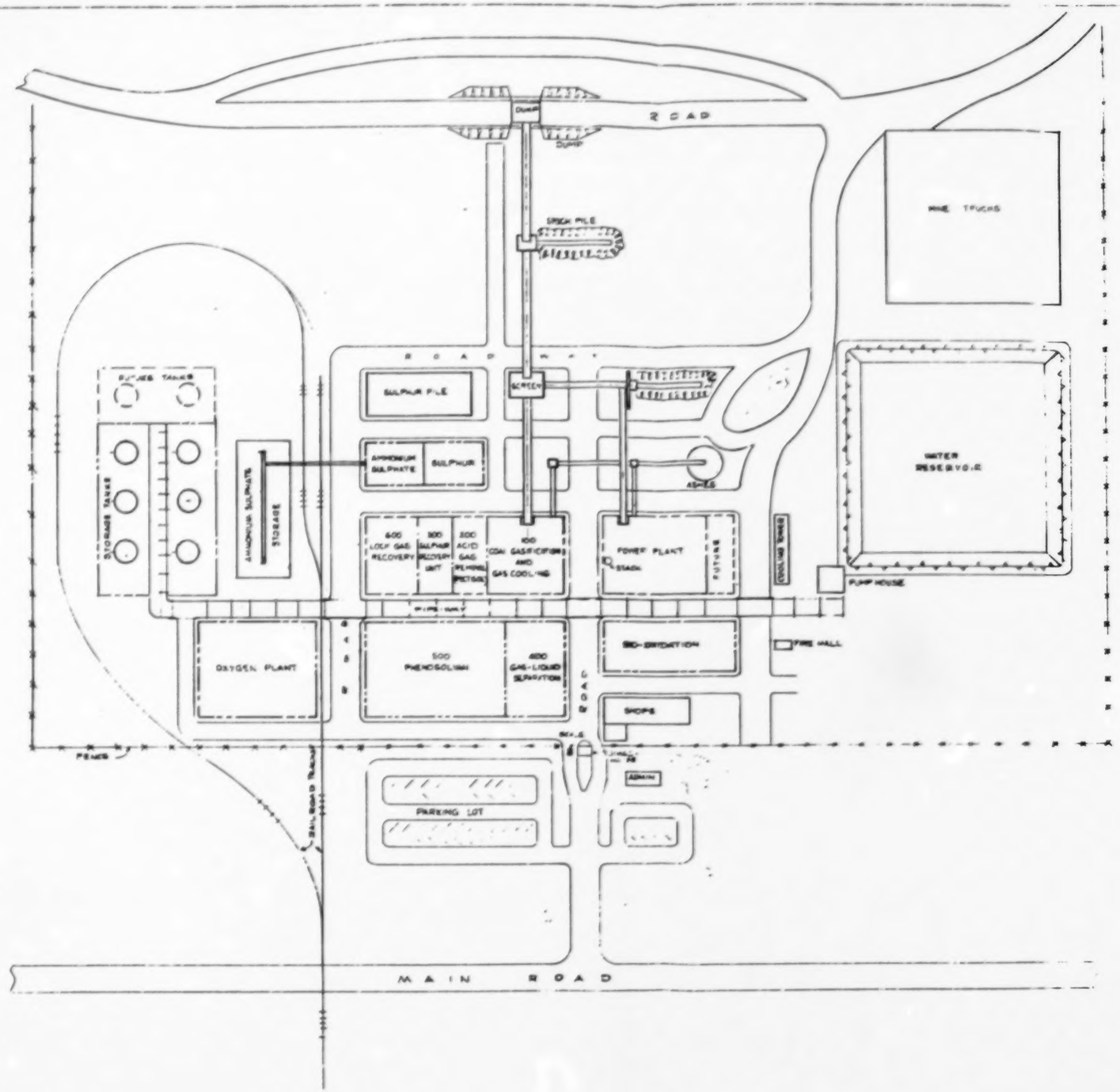
SNC TOTTRUP

SHELL CANADA RESOURCES LTD

COAL GASIFICATION
BLOCK FLOW DIAGRAM

6/mar/77

0249 7500 4500 11



REFERENCE DWG

4500-11 BLOCK FLOW DIAGRAM

SNC TOTTRUP

SHELL CANADA RESOURCES LTD

COAL GASIFICATION

PLOT PLAN

1:2500

6/30/77

0249 4500 12